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(54) **SIMULTANEOUS HIGH EFFICIENCY  
CAPTURE OF CO<sub>2</sub> AND H<sub>2</sub>S FROM  
PRESSURIZED GAS**

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**B01D 53/52** (2006.01)

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**C10L 3/10** (2006.01)

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3/104; Y02C 10/04; Y02C 10/06; Y02C  
20/20

See application file for complete search history.

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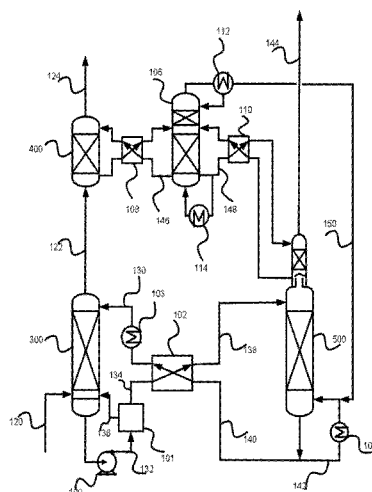
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(57) **ABSTRACT**

Low-cost and energy-efficient CO<sub>2</sub> and H<sub>2</sub>S capture is provided obtaining greater than 99.9% capture efficiency from pressurized gas. The acid species are captured in an ammonia solution, which is then regenerated by stripping the absorbed species. The solution can capture as much as 330 grams of CO<sub>2</sub> and H<sub>2</sub>S per 1000 gram of water and when regenerated it produces pure pressurized acid gas containing more than 99.7% CO<sub>2</sub> and H<sub>2</sub>S. The absorption of the acid species is accomplished in two absorbers in-series, each having multiple stages. More than 95% of the acid species are captured in the first absorber and the balance is captured in the second absorber to below 10 ppm concentration in the outlet gas. The two absorbers operate at temperatures ranging from 20-70 degrees Celsius. The two absorbers and the main stripper of the alkaline solution operate at similar pressures ranging from 5-200 bara.

**12 Claims, 5 Drawing Sheets**



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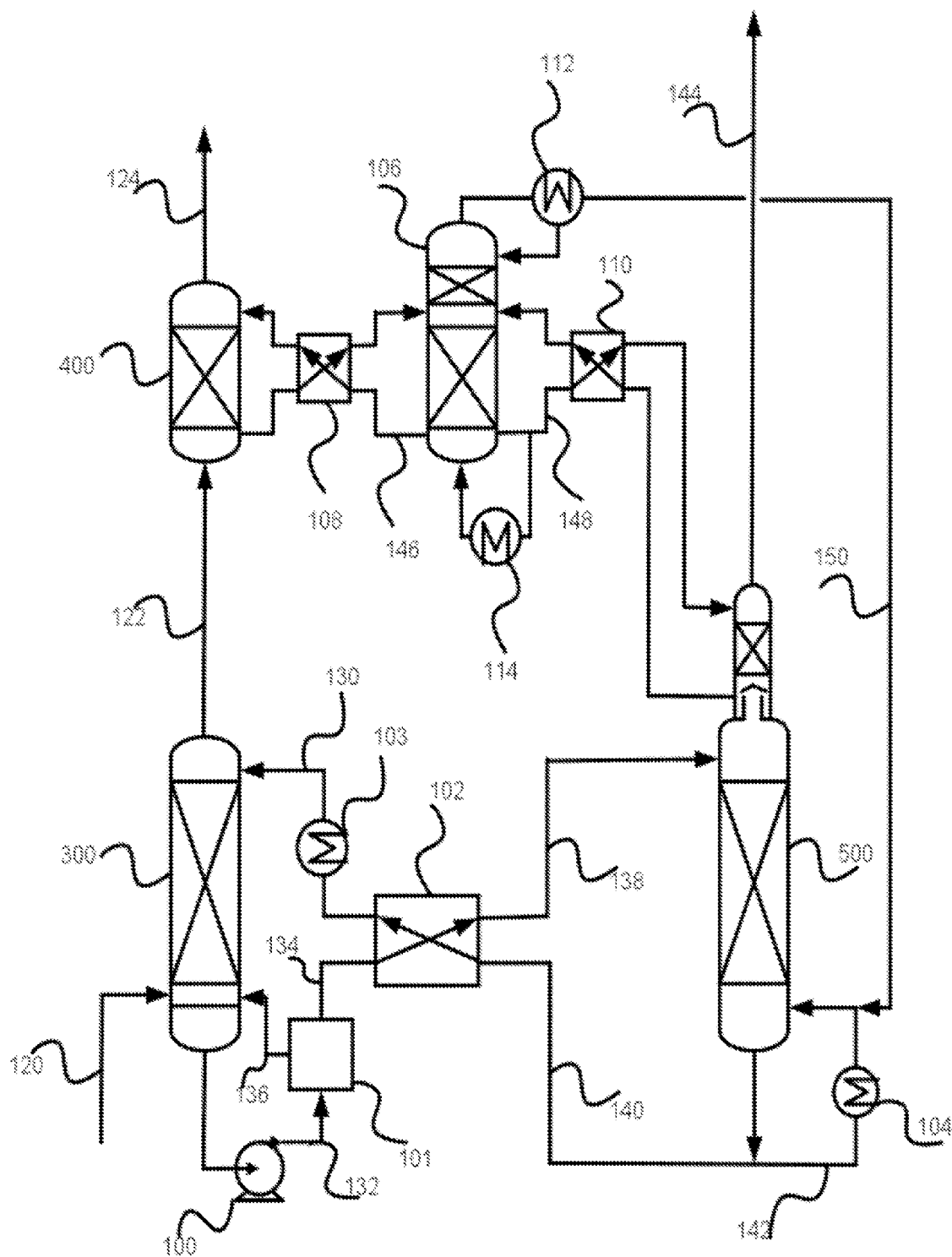


Fig. 1

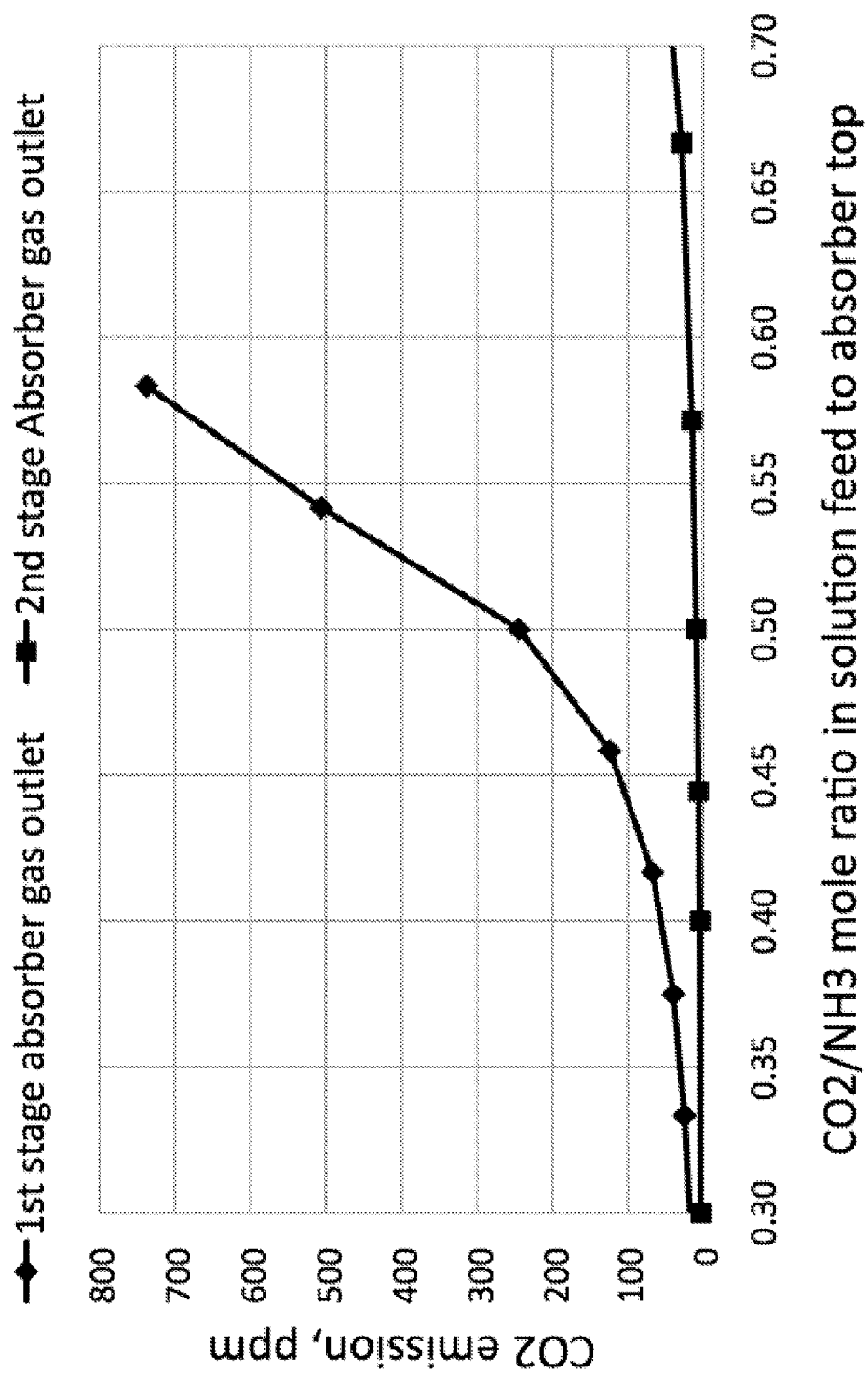


Fig. 2

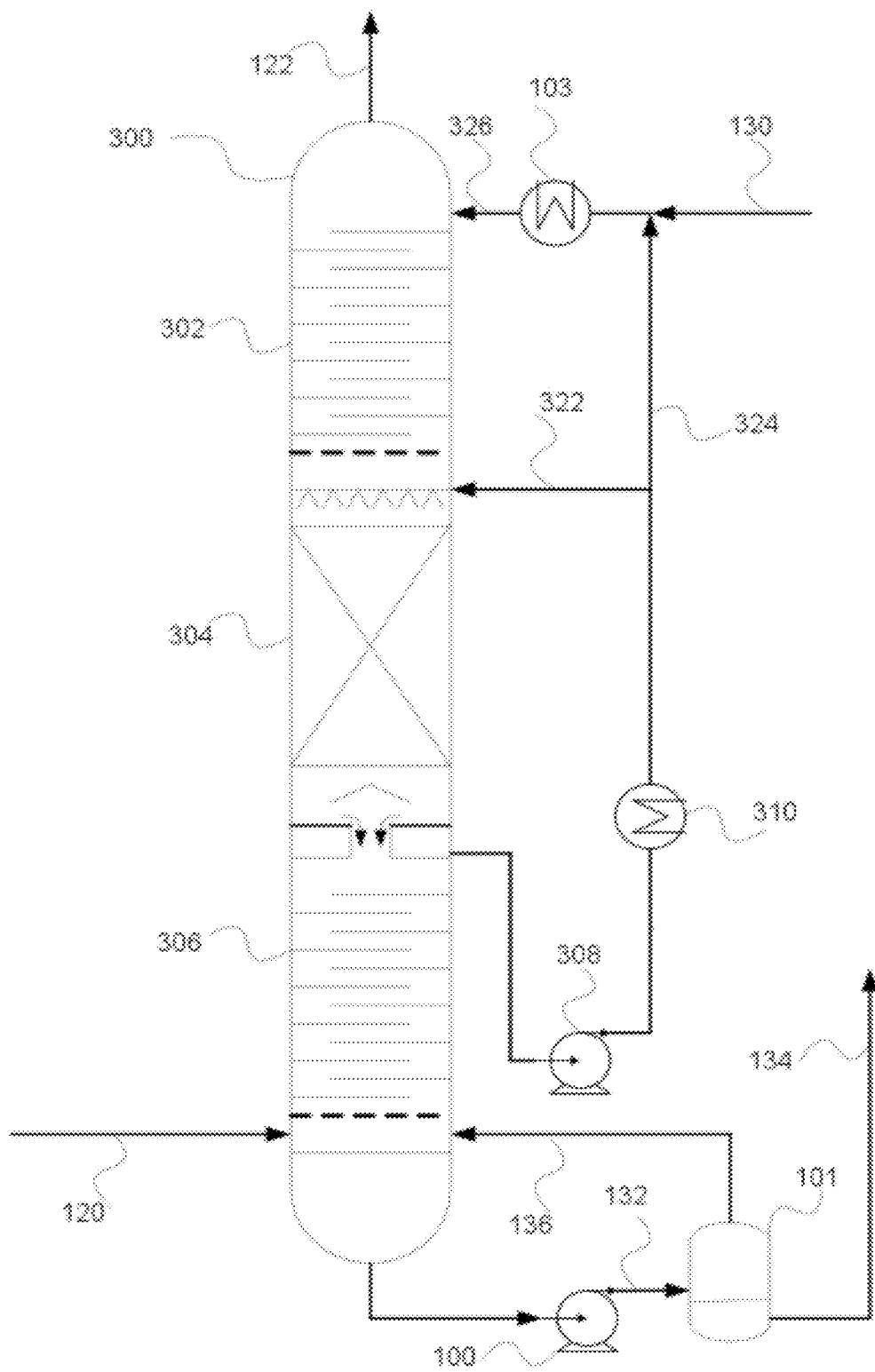


Fig. 3

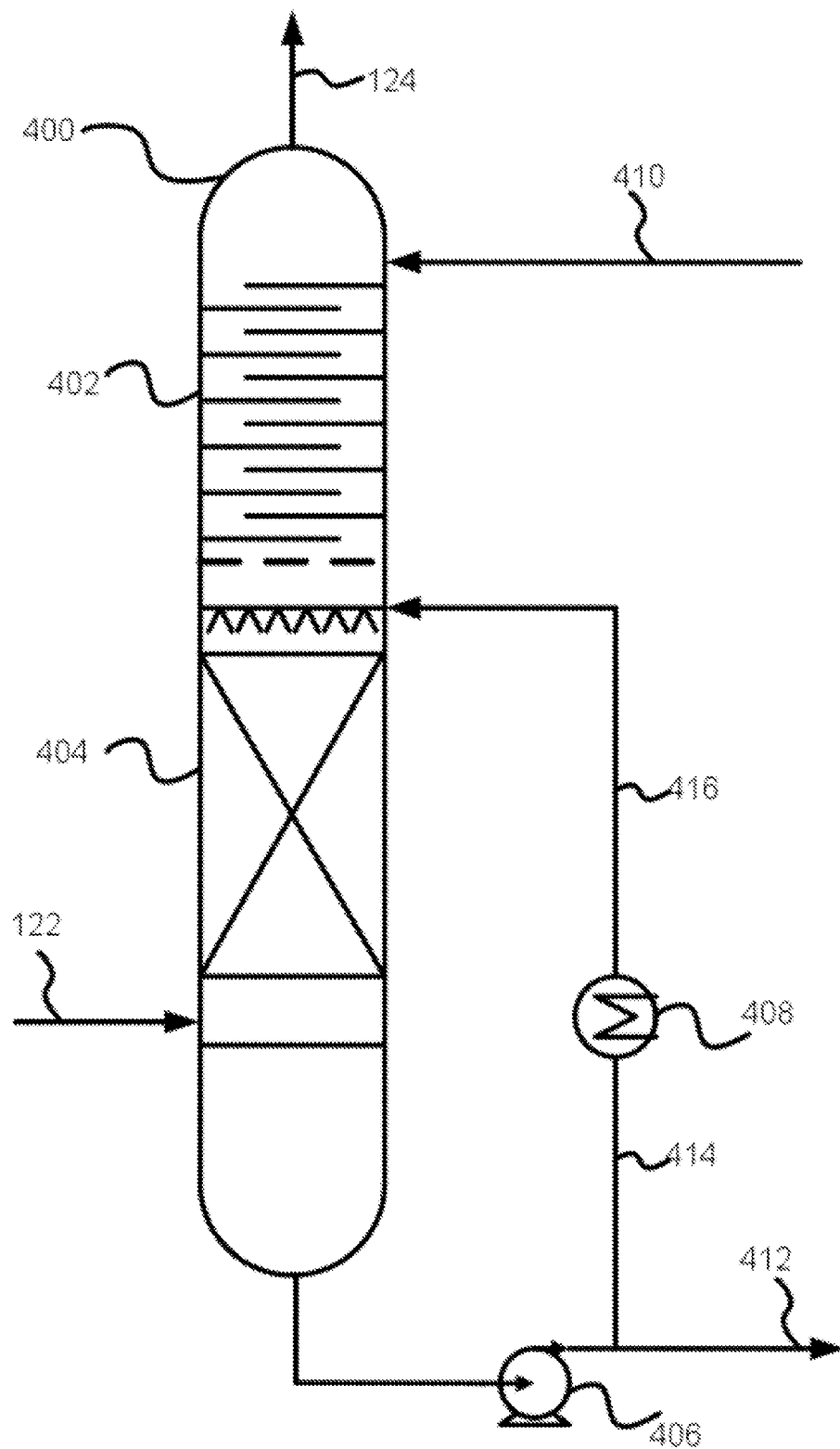


Fig. 4

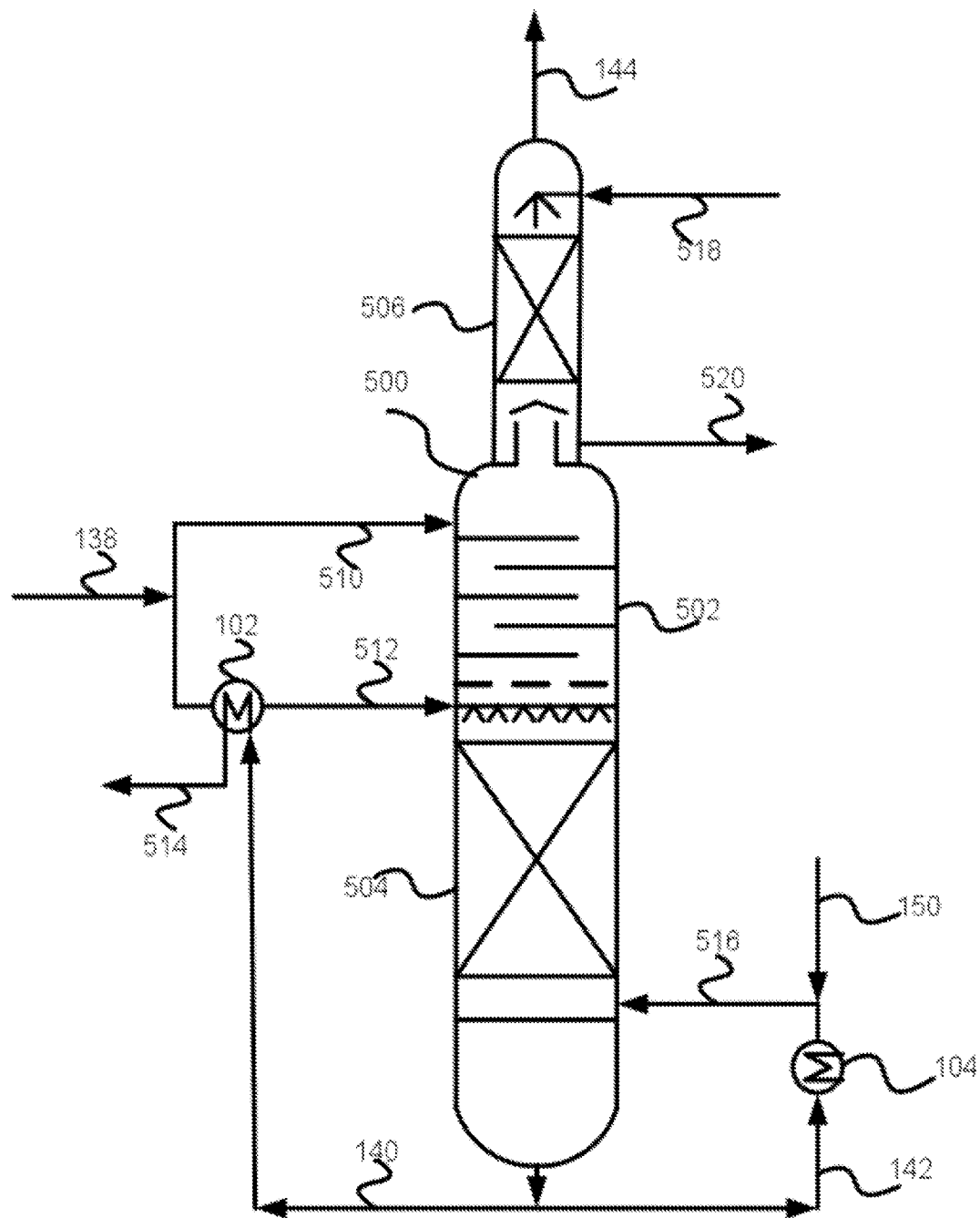


Fig. 5

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# **SIMULTANEOUS HIGH EFFICIENCY CAPTURE OF CO<sub>2</sub> AND H<sub>2</sub>S FROM PRESSURIZED GAS**

## **CROSS-REFERENCE TO RELATED APPLICATIONS**

This application is a 371 of PCT/US2012/057205 filed on Sep. 26, 2012. PCT/US2012/057205 claims the benefit of 61/539,144 filed on Sep. 26, 2011.

## **STATEMENT OF GOVERNMENT SPONSORED SUPPORT**

This invention was made with Government support under grant no. DE-FE0000896 awarded by the Department of Energy. The Government has certain rights in this invention.

## **FIELD OF THE INVENTION**

The invention relates to methods and systems for high efficiency capture of acid species, mainly CO<sub>2</sub> and H<sub>2</sub>S, from pressurized gases in alkaline solution containing ammonia and regeneration of the solution by stripping the absorbed acid species.

## **BACKGROUND OF THE INVENTION**

Capturing H<sub>2</sub>S and CO<sub>2</sub> from gases is important in many industrial processes where the H<sub>2</sub>S and CO<sub>2</sub> are contaminants and have to be removed before further processing. Capturing the H<sub>2</sub>S and CO<sub>2</sub> is also important for environmental reasons where H<sub>2</sub>S, before or after its combustion, contributes to the formation of acid rain and CO<sub>2</sub> is associated with global warming.

There are many commercial chemical and physical processes for capturing H<sub>2</sub>S and CO<sub>2</sub> from pressurized gas. Chemical processes include absorbents such as amine based processes, the Benfield process using potassium carbonate and many more. Physical processes include the Selexol process, the methanol based Rectisol process and more. These processes are typically expensive and require significant input of heat and electricity. In addition, most of available commercial processes can only capture small amount of CO<sub>2</sub> and H<sub>2</sub>S per unit volume of absorbent, typically in the 30-60 grams/liter and thus requiring the pumping and circulation of large volumes of solutions and making the reactors, pumps, pipes, heat exchangers large and expensive. In many of these processes the capture efficiency of CO<sub>2</sub> and H<sub>2</sub>S is relatively low and requires polishing steps downstream. Furthermore, another concern is that the stripped acid species are at low pressure and require high cost and energy intensive compression. Also, higher pressure gas results in higher solubility of non-acidic species such as H<sub>2</sub>, CO and CH<sub>4</sub> in the absorber outlet solution. As a result, the stripped acid gas contains H<sub>2</sub>, CO and CH<sub>4</sub> in concentrations that require further cleaning treatment and also results in losses of valuable matter.

There is a need in the art for a dramatically improved system and process for capturing H<sub>2</sub>S and CO<sub>2</sub> and reduce its cost. The present invention addresses this need.

## **SUMMARY OF THE INVENTION**

The present invention provides a process and system integrating very high capture efficiency typically greater than 99% and potentially as high as 99.99%, high acid gas

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loading in the range of 100-330 grams per 1000 grams of water (3-7 times higher loading than commercially available technologies), produce acid gas at pressure in the range of 5-200 bara and containing more than 99.7% CO<sub>2</sub> and H<sub>2</sub>S, less than 0.3% moisture and practically no non-acidic species such as H<sub>2</sub>, CO and CH<sub>4</sub>. The process consumes less than half the energy, combined heat and power than any commercial process. It utilizes a low cost ammonia reagent which is non-degradable and produces no harmful waste stream. The integrated system of the invention reduces the cost of CO<sub>2</sub> and H<sub>2</sub>S capture to less than half the cost of state of the art technologies. The high efficiency and low cost of the process is enabled by multi-stage absorption system with multiple stages each designed and operated under conditions that optimized the system performance.

Embodiments of the invention capture acid gases, mainly CO<sub>2</sub> and H<sub>2</sub>S, from pressurized gas streams into an absorbing solution and thermally strips the CO<sub>2</sub> and H<sub>2</sub>S from the absorbing solution to produce pressurized acid gas stream. The absorbing solution is a concentrated ammonia solution containing NH<sub>3</sub>—CO<sub>2</sub>—H<sub>2</sub>O—H<sub>2</sub>S. In addition to NH<sub>3</sub> the absorbing solution may contain alkaline cations such as Na<sup>+</sup>, K<sup>+</sup> and Li<sup>+</sup>.

Embodiments of the invention include the following units.

1. A multi-stage absorber where 5-15 molal ammoniated solution captures most of the CO<sub>2</sub> and H<sub>2</sub>S from a pressurized gas stream at net CO<sub>2</sub>+H<sub>2</sub>S loading in the range of 100-330 grams per 1000 grams of water.

2. A multi-stage polishing absorber weak ammonia water solution containing 0-0.2 molal ammonia is used to capture the residual CO<sub>2</sub>+H<sub>2</sub>S in the gas. In addition, the wash solution captures ammonia entrained from the absorber.

3. A main CO<sub>2</sub>+H<sub>2</sub>S stripper where the CO<sub>2</sub> and H<sub>2</sub>S are stripped from the solution at 5-200 bara pressure to generate pure acid gas stream containing more than 99.7% CO<sub>2</sub>+H<sub>2</sub>S less than 0.3% H<sub>2</sub>O and practically no H<sub>2</sub>, CO, CH<sub>4</sub> and NH<sub>3</sub>.

4. A sour water stripper where NH<sub>3</sub>, CO<sub>2</sub> and H<sub>2</sub>S species captured in the polishing absorber are stripped from the water.

Advantages of embodiments of the invention result in much lower capital costs, energy consumption and overall operating costs than any state of the art technology for CO<sub>2</sub> and H<sub>2</sub>S captures and it could reduce the cost of unit CO<sub>2</sub> and H<sub>2</sub>S captured by more than 50%.

## **BRIEF DESCRIPTION OF THE DRAWINGS**

FIG. 1 shows a schematic of a process and system for the simultaneous high efficiency capture of CO<sub>2</sub> and H<sub>2</sub>S from pressurized gas according to an embodiment of the invention. Elements 108 and 110 are heat exchangers, and element 112 is a compressor.

FIG. 2 shows according to an embodiment of the invention data of the emission of CO<sub>2</sub> from the first stage absorber and the second stage absorber when the solution is at equilibrium with the gas.

FIG. 3 shows according to an embodiment of the invention a schematic of a three-stage first multistage absorber for the capture of CO<sub>2</sub> and H<sub>2</sub>S.

FIG. 4 shows according to an embodiment of the invention a schematic of a two-stage second multistage absorber for the capture of CO<sub>2</sub>, H<sub>2</sub>S and NH<sub>3</sub>.



FIG. 5 shows according to an embodiment of the invention a schematic of typical CO<sub>2</sub> and H<sub>2</sub>S stripper with top stage washer for NH<sub>3</sub> capture.

#### DETAILED DESCRIPTION

The present invention is a system and process for the high efficient capture of acid gases mainly CO<sub>2</sub> and H<sub>2</sub>S. The acid gas species are captured simultaneously in an alkaline solution containing ammonia or a combination of ammonia and cations such as Na<sup>+</sup>, K<sup>+</sup> and Li<sup>+</sup>.

A schematic of the system for high efficiency capture of CO<sub>2</sub> and H<sub>2</sub>S from pressurized gas stream is shown in FIG. 1. Stream 120 is a gas stream at a pressure of 5-200 bara, temperature in the range of 10-75 degrees Celsius and containing CO<sub>2</sub> or CO<sub>2</sub>+H<sub>2</sub>S. Stream 120 can be syngas from coal or petcoke gasification, syngas from fuel gas steam or auto thermal reformer, natural gas from gas wells, refinery process gas and more. The gas is typically water-saturated and its CO<sub>2</sub> concentration is 1-50% mole and H<sub>2</sub>S concentration is 0-7%. Stream 120 flows through the first stage CO<sub>2</sub> and H<sub>2</sub>S absorber 300 where more than 95% of the CO<sub>2</sub> and H<sub>2</sub>S are captured by the absorbing solution. Stream 122 contains the residual of the acid species not captured in absorber 300 and in addition, stream 122 contains NH<sub>3</sub> derived from the vapor pressure of the absorbing solution inlet to the first stage absorber 300.

Stream 122 flows to second stage absorber 400 for further cleaning. The feed absorbing solution in the second stage absorber 400 is water from sour water stripper unit 106 containing low concentration of NH<sub>3</sub>, typically in the range of 0-0.2 molal. The absorbing solution captures the residual CO<sub>2</sub> and H<sub>2</sub>S to very low levels and in addition it captures entrained ammonia from the first stage absorber. The system is designed in such a way that the gas stream from the first stage absorber 300 contains ammonia to CO<sub>2</sub> plus H<sub>2</sub>S mole ratio smaller than 0.4. The clean gas outlet from the second stage absorber 400, stream 124, contains less than 10 ppm ammonia, less than 10 ppm CO<sub>2</sub> and less than 1 ppm H<sub>2</sub>S.

The inlet absorption solution in the absorber, stream 130, is an ammoniated solution containing 8-15 molal NH<sub>3</sub> and with a mole ratio of CO<sub>2</sub>/NH<sub>3</sub> in the range of 0.2-0.4. The concentration of CO<sub>2</sub> and H<sub>2</sub>S depends on the stripper operation and it can vary depending on the specific application. Stream 130 is fed to the top of a multistage first stage absorber vessel and after absorbing the bulk of the CO<sub>2</sub> and H<sub>2</sub>S, e.g. more than 95%, it is discharged at the bottom as a CO<sub>2</sub>+H<sub>2</sub>S rich solution, stream 132, with mole ratio of CO<sub>2</sub>/NH<sub>3</sub> in the range of 0.6-0.7.

FIG. 2 shows an example of the CO<sub>2</sub> emission limit at equilibrium from the top of the first and second stage absorbers operating at 50 Bara pressure and at 35 degrees Celsius solution feed temperature. It also shows that the first stage absorber CO<sub>2</sub> emission can be as low as 100 ppm at CO<sub>2</sub>/NH<sub>3</sub> mole ratio below 0.45 and that the CO<sub>2</sub> emission from the second stage absorber is nil.

Referring back to FIG. 1, the system is typically designed with the absorber 300 and the stripper 500 operating at similar pressures and in the range of 5-200 Bara. In FIG. 1, the stripper pressure is only 1-3 bar higher than that of the absorber and low pressure drop pump 100 is required to pump the rich solution, stream 132, from the absorber through a flash chamber 101 and recuperating heat exchanger 102 to the stripper. In the flash chamber the dissolved H<sub>2</sub>, CO, CH<sub>4</sub> and other non-acidic gas species are stripped from the rich solution to form gas stream 136 that flows back to the absorber. Gas stream 136 also contains

small amount of CO<sub>2</sub>, H<sub>2</sub>S and NH<sub>3</sub>. The solubility of H<sub>2</sub>, CO and CH<sub>4</sub> in the solution at the bottom of the absorber is low and the flash chamber reduces the concentration to single digit ppm levels. As a result, the stripper gas outlet, stream 144, is practically free of H<sub>2</sub>, CO and CH<sub>4</sub>. The solution feed stream to the stripper 500, stream 138, is a heated solution with heat content that is recovered in the recuperating heat exchanger 102 from the lean solution stream 140.

Heat input to the stripper is typically in the range of 40-60 KJ per mole of acid gas stripped is delivered to the reboiler 104 by heating recycle stream 142. The ammoniated solution is chemically stable and does not degrade under the operating conditions of the stripper. As a result, the heat source in the reboiler is not limited to using condensing steam, but it can also use other sources of heat such as hot syngas, hot flue gas, hot oil from solar collectors, hot brines etc.

Stream 140 is a hot, typically in the range of 150-250 degrees Celsius, and CO<sub>2</sub>/H<sub>2</sub>S lean solution from the stripper. It is cooled in the recuperating heat exchanger 102 while heating the rich solution stream 134. Further cooling of the lean solution is provided in heat exchanger 103. Typical temperature of the feed to the absorber, stream 130, when using cooling tower water for heat sink in heat exchanger 103, is 20-40 degrees Celsius.

The stripper 500 is designed in such a way that its temperature at the gas outlet is lower than 40 degrees Celsius and typically in the range of 20-40 degrees Celsius. As a result water and ammonia concentration in the outlet gas stream is low, corresponding to their vapor pressure over the inlet solution to stage 506 in FIG. 5. For example, when the temperature at the stripper gas outlet is 40 degrees Celsius and the stripper operates at 50 Bara the moisture content of the acid gas from the stripper is about 0.15% and ammonia concentration is about 1 ppm. The acid gas stream 144 from the stripper contains more than 99.7% CO<sub>2</sub> and H<sub>2</sub>S, less than 0.3% water vapor and practically no NH<sub>3</sub>, H<sub>2</sub>, CO and CH<sub>4</sub>.

The water from second stage absorber 400 and from the top of the stripper 500 contains NH<sub>3</sub>, CO<sub>2</sub> and H<sub>2</sub>S captured from the pressurized gas stream 122 and from the product acid gas stream 144. The water is sent to a sour water treatment system where heat is provided through reboiler 114 to generate water containing low concentration of ammonia in the range of 0-0.2 molal. The treated water from stripper 106 is re-used and is sent back to the second stage absorber, stream 146, and to the top of the main stripper, stream 148. The gas from the sour water stripper 106 containing CO<sub>2</sub>, H<sub>2</sub>S, NH<sub>3</sub> and water vapor, stream 150, is sent to the bottom of the stripper 500. Depending on the relative operating pressures of the sour stripper 106 and the main stripper 500 a compressor may be used to push gas stream 150 to the main stripper.

#### First Stage Absorber Vessel and System

The first stage absorber and system is a multistage vessel 300 with at least two absorption stages each designed to achieve optimal results. A schematic of a three-stage absorber designed for high efficiency capture of CO<sub>2</sub> and H<sub>2</sub>S is shown in FIG. 3.

Feed gas, stream 120, containing CO<sub>2</sub> and H<sub>2</sub>S is injected to the bottom of the absorber, stage 306, and it flows upwards through the absorber stages 304 and 302 to exit as clean outlet gas at the top, stream 122. The solution fed to the top of the absorber, stream 326, is a mix of lean solution from the stripper, stream 130, and semi-rich solution from the second stage, stream 324. The resultant stream has

CO<sub>2</sub>/NH<sub>3</sub> mole ratio of 0.3-0.4 a ratio which is designed to optimize the capture of CO<sub>2</sub> and H<sub>2</sub>S while minimizing ammonia emission from the absorber. The mixed gas stream is cooled in heat exchanger **103** to below 40 degrees Celsius before it is fed to the top of the absorber.

For example, the equilibrium gas concentration above 12 molal NH<sub>3</sub> solution at 35 degrees Celsius and 50 bara and containing CO<sub>2</sub>/NH<sub>3</sub> mole ratio of 0.33 is about 4,000 ppm NH<sub>3</sub> and less than 100 ppm each for CO<sub>2</sub> and H<sub>2</sub>S. The absorber gas outlet can be designed to achieve ammonia equilibrium concentration of 4000 ppm, CO<sub>2</sub> concentration above equilibrium and less than 1,500 ppm and H<sub>2</sub>S concentration of 100 ppm. It is important to keep the acid to ammonia mole ratio in the gas at above 0.4 so that after capturing all the residual species from the gas the second stage absorber solution is highly alkaline with CO<sub>2</sub>/NH<sub>3</sub> smaller than 0.4 so that it can capture all the residual CO<sub>2</sub> and H<sub>2</sub>S.

The top stage absorber in FIG. **3** is shown as a tray tower, but it can also be a packed tower or other gas-liquid contacting device. Also, the stage can operate as a once through liquid as shown in FIG. **3** or it can have a recycle. Due to the low CO<sub>2</sub>/NH<sub>3</sub> mole ratio of the solution, 0.3-0.4 in the top stage **302**, the absorption rate of CO<sub>2</sub> and H<sub>2</sub>S into the solution is high. At the liquid outlet from the top absorber stage the solution is 3-15 degrees Celsius warmer than the feed solution **326** due to the heat of reaction of the CO<sub>2</sub> and H<sub>2</sub>S absorption.

The solution from the top stage is fed to the middle stage **304** where it is mixed with cooled recycle solution, stream **322**. FIG. **3** shows the middle absorber stage **304** as a packed tower with recycle. Other gas-liquid contacting devices can be applied as well. The recycle stream in the middle stage, stream **322**, is designed to increase liquid flow in the stage, to increase the rate of mass transfer and to control the stage temperature. Heat exchanger **310** removes excess heat of reaction from the recycled solution and it prevents overheating of the solution maintaining the temperature at 40-60 degrees Celsius. The recycle of the solution is done by recycle pump **308**. Excess solution from the middle stage flows to the bottom stage.

The bottom stage **306** absorber is designed to produce CO<sub>2</sub> and H<sub>2</sub>S rich solution and to maximize the CO<sub>2</sub> and H<sub>2</sub>S loading of the solution. Depending on the partial pressure of CO<sub>2</sub> and H<sub>2</sub>S in the gas feed, stream **120**, and on the design characteristics of the stage, i.e. height of the stage, gas velocity, type of packing and operating temperature the outlet solution from the absorber can have as high as 0.7 CO<sub>2</sub> to NH<sub>3</sub> mole ratio and net loading, the difference in CO<sub>2</sub> and H<sub>2</sub>S content between the solution inlet to the absorber, stream **130**, and the solution outlet from the absorber, stream **134**, as high as 330 grams per 1000 grams of water or 7.5 molal of CO<sub>2</sub>.

The high ionic strength and the high CO<sub>2</sub> loading of the solution at the bottom stage of the absorber may result in the precipitation of crystals of ammonium bicarbonate. For example, solution containing 12 molal of ammonia and having acid-to-NH<sub>3</sub> mole ratio of 0.7 should be at temperature greater than 60 degrees Celsius to prevent solids precipitation. As a result, the cooling of the middle stage absorber in heat exchanger **310** is controlled in such a way that the temperature in the bottom stage is 3-5 degrees Celsius higher than the precipitation temperature of solids.

Depending on the absorber pressure, H<sub>2</sub>, CO, CH<sub>4</sub> can be physically absorbed in the solution. To eliminate practically all physically dissolved species from the solution so that loss of valuable species is eliminated and the acid gas from the

main stripper contains only CO<sub>2</sub> and H<sub>2</sub>S a flash chamber **101** is installed at the solution outlet from the absorber. The outlet solution from the absorber, stream **132**, is flashed into the flash vessel optionally after heating the solution by 3-10 degrees Celsius. The physically dissolved species in the solution are flashed out of the solution and is sent back, stream **136**, to the bottom of the absorber.

#### Second Stage Absorber Vessel and System

The second stage absorber and system is designed to produce gas containing low concentration of CO<sub>2</sub>, H<sub>2</sub>S and NH<sub>3</sub> all in the less than 10 ppm level. The second stage absorber, Vessel **400** has at least two absorption stages. A schematic of a two-stage second absorber is shown in FIG. **4**.

The inlet gas stream to the second stage absorber is stream **122**, which is the outlet stream from the first stage absorber. It contains residual CO<sub>2</sub> and H<sub>2</sub>S and in addition it contains NH<sub>3</sub> that evaporated from the ammoniated solution in the absorber. The first stage absorber is controlled in such a way that the CO<sub>2</sub> plus H<sub>2</sub>S to NH<sub>3</sub> mole ratio in gas stream **122** is less than 0.4 and as a result, the solution in the second stage absorber is highly alkaline and capable of removing residual CO<sub>2</sub> and H<sub>2</sub>S from the gas.

In addition to low CO<sub>2</sub>, H<sub>2</sub>S and NH<sub>3</sub> emission the second stage absorber is designed to minimize the use of water which is achieved by minimizing the NH<sub>3</sub> emission from the first stage absorber and by producing high ammonia concentration bleed stream in the range of 1-6 molal. In the example shown in FIG. **4**, the absorber has 2 stages. The bottom stage **404** is a packed tower and it utilizes recycle pump **406** to recycle solution within the absorbing stage, stream **414**, and to discharge excess solution, stream **412**, from the system and sending it to the sour water stripper. The recycle solution, stream **414**, is cooled in heat exchanger **408** to 3-10 degrees Celsius above the cooling water temperature to produce cooled solution, stream **416**. Stream **416** is fed to the top of the stage **404** and is mixed with solution from the top stage **402**. The bottom stage operates at molality in the range of 1-6 and it captures most of the ammonia from the gas as well as the most of the residual CO<sub>2</sub> and H<sub>2</sub>S in the gas stream.

The top stage of the second stage absorber **402** in FIG. **4** is a counter flow tray tower where water containing ammonia concentration in the range of 0-0.2 molal from the sour water stripper, stream **410**, is cooled to 3-10 degrees Celsius above cooling water temperature. The cooled water is fed to the top of the absorber and flow downwards through the trays and it captures residual NH<sub>3</sub>, H<sub>2</sub>S and CO<sub>2</sub> from the gas to below 10 ppm mole concentrations. Different type of packing or trays may be used in stages **402** and **404**.

#### Main CO<sub>2</sub> and H<sub>2</sub>S Stripper

The main CO<sub>2</sub> and H<sub>2</sub>S stripper is designed to strip CO<sub>2</sub> and H<sub>2</sub>S from the rich solution produced in the absorber (CO<sub>2</sub>/NH<sub>3</sub>=0.6-0.7 mole ratio) and to convert it to lean solution (CO<sub>2</sub>/NH<sub>3</sub>=0.2-0.3 mole ratio). The stripper operates at pressure in the range of 5-200 Bara and typically at a pressure close to the pressure of the absorber. The CO<sub>2</sub> and H<sub>2</sub>S stripping is done with practically no loss of NH<sub>3</sub> from the system.

A schematic of typical main CO<sub>2</sub> and H<sub>2</sub>S stripper **500** is shown in FIG. **5**. Rich solution from the absorber, stream **138**, is split to two. The smaller stream **510** is relatively cold and CO<sub>2</sub>-rich solution and it flows to the top of the stage **502**. The temperature of the stream **510** solution is typically at 60-80 degrees Celsius, to avoid the precipitation of ammonium bicarbonate. The solution flows downwards in stage **502** through a series of trays cooling the rising acid gas

and capturing and condensing the ammonia and the water vapor in the gas. The heat of reaction and condensation and the sensible heat of the rising acid gas increase the solution temperature on its way down and it recovers the heat that otherwise would be lost.

The solution from stage **502** liquid outlet is mixed with stream **512**, the main rich stream from the absorber, which is heated in a recuperating heat exchanger before entering the stripper.

The bottom stage of the stripper, stage **504**, is typically a packed tower where hot gas from the reboiler **104**, at typical temperature in the range of 150-200 degrees Celsius or higher and containing CO<sub>2</sub>, H<sub>2</sub>S, NH<sub>3</sub> and H<sub>2</sub>O, flows upwards counter-currently to the rich feed solution. Heat and mass transfer occurs in the packed section of the stripper where the less volatile species from the gas, H<sub>2</sub>O and NH<sub>3</sub> vapor are cooled and condensed in the solution, while the more volatile species in the solution, CO<sub>2</sub> and H<sub>2</sub>S, evaporate into the gas phase. As a result, the rising gas becomes richer in CO<sub>2</sub> and H<sub>2</sub>S and leaner in NH<sub>3</sub> and H<sub>2</sub>O. Further enrichment of the gas in CO<sub>2</sub> and H<sub>2</sub>S occurs in stage **502** of the stripper.

Heat is provided to the stripper in the reboiler **104**. The heat source to the reboiler can be any hot stream such as steam, syngas, flue gas and even heated oil from solar collectors. Stream **142** is a feed solution to the reboiler and stream **516** is a two phase stream from the reboiler. The gas phase in stream **516** contains the gas species that evaporated in the reboiler as well as gas species from the sour water stripper stream **150**.

Hot lean solution, stream **140**, is withdrawn from the bottom of the stripper and sent to the recuperating heat exchanger **102** to cool the solution, stream **514**, and to recover its heat. In a system where the stripper pressure is higher than the absorber pressure a pump is installed to pump the rich solution to the stripper. In a system where the stripper pressure is lower than the absorber pressure the pump is installed to pump the lean solution to the absorber.

A wash stage **506** is installed at the top of the stripper and is designed to capture all the ammonia from the gas stream and to further reduce the moisture content of the gas stream. Stage **506** is a packed or tray tower where cooled water from the sour water stripper, stream **518**, is fed to the top and flows downwards counter currently to the rising acid gas. The high partial pressure of the CO<sub>2</sub> in the acid gas results in high concentration of dissolved CO<sub>2</sub> in the solution and enhances the capture of NH<sub>3</sub>. The outlet solution from stage **506**, stream **520** contains practically all the ammonia that enters the stage in the gas phase.

The outlet gas stream **144** from the top of the main stripper is pure CO<sub>2</sub> and H<sub>2</sub>S stream except for 0.1-0.3% of water vapor that can be easily removed downstream. Stream **144** contains practically no H<sub>2</sub>, CO, CH<sub>4</sub> and other physically absorbed species from the absorber.

#### Sour Water Stripper

The sour water stripper collects water containing NH<sub>3</sub>, CO<sub>2</sub> and H<sub>2</sub>S from the second stage absorber **400** in FIG. **4** and from the top stage of the main stripper **506** in FIG. **5**. The sour water stripper is a conventional thermal stripper preferably operating at high pressure in such a way that the stripped gas **150** flows to the main acid gas stripper **500** by the pressure difference between the two vessels. Otherwise, a compressor is used to send the gas to the main stripper.

What is claimed is:

1. A system for a multi-staged capture of CO<sub>2</sub> and H<sub>2</sub>S from a pressurized gas stream producing a clean pressurized gas stream containing less than 1 ppm H<sub>2</sub>S and less than 10

ppm CO<sub>2</sub> and achieving greater than 99.9% capture efficiency, wherein the pressure in said pressurized gas stream and said clean pressurized gas stream is above atmospheric pressure, comprising:

- (a) an alkaline solution containing 5-15 molal NH<sub>3</sub>;
- (b) a first multistage absorber for absorbing CO<sub>2</sub> and H<sub>2</sub>S from said pressurized gas stream combined with said alkaline solution, wherein greater than 95% of said CO<sub>2</sub> and H<sub>2</sub>S content of said pressurized gas stream is captured in said alkaline solution, wherein a net loading of 110-330 grams of said CO<sub>2</sub> and H<sub>2</sub>S per 1000 grams of water is absorbed in said alkaline solution;
- (c) a second multistage polishing absorber using water for capturing the residual of CO<sub>2</sub> and H<sub>2</sub>S in said pressurized gas stream entrained from said first multistage absorber and NH<sub>3</sub> entrained from said first multistage absorber resulting in a gas outlet stream from said second multistage polishing absorber containing less than 10 ppm CO<sub>2</sub> and NH<sub>3</sub> and less than 1 ppm H<sub>2</sub>S;
- (d) a main stripper for stripping said CO<sub>2</sub> and H<sub>2</sub>S absorbed by said first multistage absorber in said alkaline solution, resulting in:
  - (i) a pure gas stream exhausted containing greater than 99.7% CO<sub>2</sub>+H<sub>2</sub>S, less than 0.3% water vapor and less than 10 ppm of each of H<sub>2</sub>, CO, NH<sub>3</sub> and CH<sub>4</sub>, and
  - (ii) said alkaline solution reduced of said stripped CO<sub>2</sub> and H<sub>2</sub>S which is sent back from said main stripper to said first multistage absorber,
- (e) a sour water stripper for stripping said CO<sub>2</sub>, H<sub>2</sub>S and NH<sub>3</sub> absorbed by said second multistage polishing absorber in said water, and said water stripped of said CO<sub>2</sub>, H<sub>2</sub>S and NH<sub>3</sub> is sent back from said sour water stripper to said second multistage polishing absorber.

2. The system as set forth in claim 1, wherein said pressurized gas stream comprises an H<sub>2</sub>S concentration in the range of 0-7% mole and a CO<sub>2</sub> concentration in the range of 1-50% mole.

3. The system as set forth in claim 1, wherein said alkaline solution further comprises alkaline cations, wherein said alkaline cations are Na<sup>+</sup>, K<sup>+</sup>, Li<sup>+</sup>, or a combination thereof.

4. The system as set forth in claim 1, wherein the said pressurized gas stream is at a pressure in the range of 5-200 bara, wherein said first multistage absorber and said second multistage polishing absorber operate at a pressure in the range of 5-200 bara, and wherein said main stripper operates at a pressure in the range of 5-200 bara.

5. The system as set forth in claim 1, wherein said first multistage absorber comprises at least two absorption stages, wherein each of said absorption stages is a gas-liquid contact vessel.

6. The system as set forth in claim 5, wherein one of said at least two absorption stages comprises: (i) said alkaline solution having a NH<sub>3</sub>/CO<sub>2</sub> mole ratio in the range of 0.2-0.4, and (ii) an operating temperature in the range of 20-40 degrees Celsius.

7. The system as set forth in claim 5, wherein one of said at least two absorption stages comprises: (i) said alkaline solution having a NH<sub>3</sub>/CO<sub>2</sub> mole ratio in the range of 0.3-0.55, and (ii) an operating temperature in the range of 35-60 degrees Celsius.

8. The system as set forth in claim 5, wherein one of said at least two absorption stages comprises: (i) said alkaline solution having a NH<sub>3</sub>/CO<sub>2</sub> mole ratio in the range of 0.5-0.7, and (ii) an operating temperature in the range of 50-70 degrees Celsius.

9. The system as set forth in claim 1, wherein said second multistage polishing absorber comprises at least two absorption stages to reduce  $\text{CO}_2$  and  $\text{H}_2\text{S}$  emission to ppm level concentrations.

10. The system as set forth in claim 9, wherein one of said at least two absorption stages comprises: (i) a water-ammonia solution with an  $\text{NH}_3$  concentration in the range of 0-0.2 molal, and (ii) a  $\text{CO}_2/\text{NH}_3$  mole ratio in said water of less than 0.5.

11. The system as set forth in claim 9, wherein one of said at least two absorption stages comprises: (i) an  $\text{NH}_3$  concentration in the range of 1-6 molal, and (ii) a  $\text{CO}_2/\text{NH}_3$  mole ratio in said water of less than 0.5.

12. The system as set forth in claim 1, wherein said main stripper further comprises an integrated  $\text{NH}_3$  washing stage.

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